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CYANAMID

GENERATION OF HYDROGEN FOR FUEL CELLS IN SUBMARINE PROPULSION

Processes Involving Decomposition of Ammonia

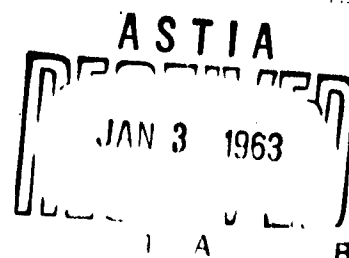
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INTRODUCTION

INTRODUCTION

Fuel cells are under study as sources of electrical power for submarines. At present only the hydrogen-oxygen cell is sufficiently advanced for consideration. The oxygen for this cell could be stored as liquid aboard the submarine, but methods for handling liquid hydrogen under similarly practical conditions are not now available. Therefore, a hydrogen plant must be installed in the submarine.

Several processes are in industrial use for production of hydrogen, using various raw materials and methods. Any of these can be adapted for installation on a submarine. For use on a submerged submarine, however, they require drastic modification.

Waste gases which are normally produced by these processes must not be vented, except when snorkeling. Equipment noise, size and response time must be minimized. All of the equipment must be capable of withstanding severe shock loads and of operating properly under all maneuvering positions of the submarine.

To meet these and other requirements, we have prepared alternate process designs for using the decomposition of ^{NH₃} ammonia as a source of hydrogen. Our research laboratories have conducted experiments to provide certain basic kinetic data and catalyst information; our high speed digital computers have been used to develop mathematical models; and our drafting facilities have constructed space models for our designs.

Our study leads to the conclusion that one of the alternates is immediately available for firm design and installation. A second one may be developed readily as a slight improvement, but a third one which requires development by pilot plant investigation promises to become the

optimum process. This report describes our designs for the three alternatives. It also includes the assumptions on which the designs are based and a suggested program to select and install the most practical process.

ABSTRACT

ABSTRACT

Three raw materials for production of hydrogen are compared on a theoretical basis. The reasons for selection of ammonia for further study are set forth.

Two terms should be defined at the beginning. We refer to "usable hydrogen" as hydrogen fed to the cells less hydrogen purged from them. "Net real hydrogen" is the flow of usable hydrogen less the amount used to produce power for the hydrogen generation process itself. Using ammonia as raw material, three process alternates are treated. They differ primarily in the method used to cause dissociation of the ammonia into nitrogen and hydrogen. The specified design basis for each case is that at maximum rate the plant must produce 70 lbs./hr. of net real hydrogen. While emphasizing maximum utilization of commercially available equipment, we have allowed our design to exceed the specified basis. This overdesign can be eliminated by substituting other equipment to be found by more exhaustive equipment survey or to be ordered as special supplies for the specified rates. However, it is not considered purposeful to do so until a process is chosen subsequent to this study.

In Alternate I, the dissociation is carried out at low pressure, i.e., about 35 psig, in conventional equipment using electrical energy to supply the heat required by the endothermic reaction. In Alternate II, which is treated briefly, the dissociation is carried out at a higher pressure than is commonly used; electrical energy is the heat source as in Alternate I. In Alternate III, the heat of dissociation is supplied by combustion of ammonia in the stream of ammonia being decomposed; and both reactions go on at high pressure. All three alternates are alike in that

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the by-product nitrogen is liquefied under pressure by heat exchange with boiling oxygen, and the liquid nitrogen is then stored in place of the oxygen. No gas is vented in any alternate, and no extra storage space is required.

Each of the three alternates is taken up separately. The process, equipment, instrumentation, cost and design assumptions of Alternate I are described in detail first. While this is not the most efficient or compact process conceivable, its design does not require any extension of our present knowledge in chemical engineering. As we have designed it, some of the equipment projects about $3\frac{1}{2}$ feet out of the space allocated both fore and aft. The hydrogen holdup is less than 400 cubic feet, compared with the allowable maximum of 900 cubic feet.

Alternate II is a process of unproved efficiency. It rests on a few additional assumptions which have not yet been tested, so far as we know. Its advantage is a lower power requirement than in Alternate I.

Alternate III is the most efficient process we have planned to date. Application of this process, however, would have to be preceded by bench scale and pilot plant experimentation with various catalysts and operating conditions. The equipment volume exceeds slightly the space allowed for it by the Bureau, but does not conflict with any existing equipment shown on the drawing furnished to us.

The most important characteristics of the alternate processes may be compared as follows; other details are tabulated in the Appendix;

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<u>Alternate</u>	<u>Installed Cost</u>	<u>Submarine Volume Required (cu. ft.)</u>	<u>Net Hydrogen Available for Submarine Power (lbs./hr.)</u>	<u>Operating Requirements per lb. of net real H₂</u>		
				<u>Process KWH</u>	<u>lbs. O₂</u>	<u>lbs. NH₃</u>
I	\$800,000	1480	78.4	7.85	14.5	10.1
II		1252	70.0	6.92	13.6	9.54
III	\$515,000	1295	98.0	0.28	9.9	6.96

In this study, we have considered Alternate I as the reference design from which Alternates II and III are derived by innovations and with which they are compared in regard to relative merits. As reference design, Alternate I is aimed at immediate application by minimizing uncertainties and incorporating conventional equipment. Therefore, it offers the advantage that the design can be firmed and the equipment installed without any delay whenever a decision is reached to adopt this alternate for testing in a submarine. Alternate II reduces the parasitic power requirement by approximately 20% and reduces slightly the raw material and space requirements but introduces a number of process uncertainties which must be resolved.

Alternate III is the most attractive scheme in virtue of its low requirements for parasitic power, space, and raw material and a very low hydrogen inventory. It lends itself to optimization by advanced engineering skills including process simulation and mathematical analysis using high speed electronic computers. During this study, we have already laid the foundation for more detailed investigation of this alternate in the future. Our research laboratories have also initiated programs for establishing reaction kinetics and materials behavior as well as developing catalysts of high efficiency and optimum characteristics.

Development of the above processes may be carried out in the following steps with some overlapping:

- (1) Design Alternate I for immediate installation on a test bed or submarine or both.
- (2) Modify the Alternate I design after testing to determine the feasibility of Alternate II.
- (3) Pilot Alternate III while Alternate I is being designed. If the necessary information is developed before Alternate I can be built, modify the design and install Alternate III on the test bed or submarine or both.

However, we recognize that Alternate III is our goal and it could be achieved without step (2) of the above mentioned development program.

PROCESSES FOR GENERATION OF HYDROGEN

PROCESSES FOR GENERATION OF HYDROGEN

The major industrial processes for producing hydrogen in the United States are steam reforming of natural gas and partial oxidation of liquid hydrocarbons or natural gas. Comparatively small amounts of hydrogen are made by cracking ammonia. In the past, some military requirements have been met by reforming methanol. Coke oven gas and refinery off-gas have been used in large quantities, at locations where they are produced, as impure feeds for hydrogen purification plants.

Before discussing processes which may be applied to operations inside a submarine, we should point out that methods of handling liquid hydrogen are now under intensive development. It is possible that these methods will improve sufficiently to make it useful in submarines as fast as a generation process can be developed to that stage. Therefore, it is suggested that methods of storing liquid hydrogen safely outside the pressure hull be studied while processes for generating it on board are engineered.

Returning to the above processes, the use of natural gas as a raw material is ruled out because it is not generally available, either as a gas or as liquefied methane. Partial oxidation of some liquid hydrocarbons may lead to the production of soot, which presents a disposal problem. Coke oven gas and refinery off-gas are, of course, unavailable to submarine application.

Dissociation of ammonia and reforming of methanol remain. Recently processes have also been developed for reforming light hydrocarbons auto-thermally, i.e., by oxidizing part of the hydrocarbon in the reaction feed to supply the heat required to reform the rest.

These raw materials may be compared first by visualizing ideal thermodynamic processes. The raw materials are liquid oxygen and the liquid hydrogen compounds, such as ammonia, naphtha or methanol. Where reforming is involved, water is recycled from the fuel cell. The end products are taken to be liquid water, carbon dioxide and nitrogen at the conditions of storage or disposal.

On this ideal basis the above raw materials may be compared as follows:

	Partial Oxidation of Naphtha*	LPG*	Reforming of Methanol	Dissociation of Ammonia
Raw Material Usage				
lbs./lb. usable H ₂	2.78	2.71	5.94	6.67
gals./lb. usable H ₂	0.50	0.56	0.90	1.35
Oxygen Usage				
lbs./lb. usable H ₂	9.82	9.70	8.88	9.41
gals./lb. usable H ₂	1.03	1.02	0.93	0.99
Liquid CO ₂ Produced				
lbs./lb. usable H ₂	8.55	8.00	8.14	
gals./lb. usable H ₂	1.01	0.95	0.96	
Liquid N ₂ Produced				
lbs./lb. usable H ₂				5.49
gals./lb. usable H ₂				0.82
Energy Produced				
KWH/lb. usable H ₂	16.90	16.70	17.70	17.90
KWH/lb. raw material	6.08	6.12	2.98	2.68
KWH/lb. oxygen	1.72	1.72	1.99	1.90

* Using n-hexane and n-butane for naphtha and LPG respectively.

Under perfect conditions, the hydrocarbons yield more energy per pound of raw materials, but less per pound of oxygen, than methanol and ammonia. The oxygen supply, which must be stored in a special container inside the pressure hull, is believed to limit the underwater capability of the vessel. Therefore storage problems favor methanol and ammonia.

As operators of reforming and ammonia plants, we consider reliability to be the most important criterion. The major threat to reliability in any catalytic process is damage to the catalyst.

Naphtha may contain poisons for reforming and shift catalyst. Oxides of carbon are formed in reforming naphtha and methanol, and these may poison fuel cell catalysts. In land-based chemical plants temporary poisoning may be corrected, but it may be fatal on board a submarine. Disposal of carbon dioxide presents some additional problems. Ammonia, on the other hand, does not contain poisons for dissociation catalyst. When dissociated alone, it does not form poisons for fuel cell catalysts.

For the above reasons we have concentrated on processes involving decomposition of ammonia.

PROCESSES UTILIZING DISSOCIATION OF AMMONIA

Alternate I - Dissociation at Low Pressure by Electrical Energy

PROCESSES UTILIZING DISSOCIATION OF AMMONIA

Alternate I - Dissociation at Low Pressure by Electrical Energy

Process Description

Most industrial dissociators are heated electrically and operate at pressures which are slightly above atmospheric. To make use of commercially available equipment, this process was designed around it. All of the equipment items have been specified by vendors or designed by us, using published data.

The following description refers to Process and Service Flowsheets FS-1-D and FS-2-D in the Appendix to this report. In overall summary, the process consists of dissociation of ammonia followed by condensation of the nitrogen produced. The resulting hydrogen stream is one feed for the fuel cell. The other feed stream is oxygen, which is vaporized by using it to supply refrigeration for the purification of the hydrogen. First the hydrogen and then the oxygen will be traced through the process.

Liquid ammonia from storage at 85° F. and 167 psia is reduced in pressure to 57 psia and vaporized at 28° F. in the Ammonia Vaporizer. The vapor is preheated against dissociated ammonia vapor in the Ammonia Preheater and is then cracked in the Ammonia Dissociator. Energy for the endothermic dissociation is supplied with electrical resistance elements.

The cracked ammonia contains less than 0.1% residual ammonia. It is used to preheat and vaporize the ammonia feed and is then chilled to reduce the volume for compression. Just before entering the Crude Hydrogen Compressor it is joined by two recycle streams which will be described later.

In the Crude Hydrogen Compressor the hydrogen-rich stream is compressed to 315 psia. It is aftercooled, chilled and scrubbed with cold water to remove residual ammonia. Then it is dried over molecular sieves and fed to the cryogenic equipment.

The Cold Exchanger uses several cold streams to precool the crude hydrogen and to condense out part of its nitrogen. Any traces of ammonia or water vapor that may pass through the scrubber and dryer are frozen out in this exchanger. Additional nitrogen is condensed in the Scrubbing Column, which is actually a partial condenser.

The bottoms product of the column is liquid nitrogen containing some dissolved hydrogen. The hydrogen is flashed off at 29.7 psia with some nitrogen and recycled to the suction of the Crude Hydrogen Compressor. The remaining liquid nitrogen is stored in a compartment of the liquid oxygen storage tank.

The overhead product of the column is 90% hydrogen, 10% nitrogen at 284 psia and -319° F. It is heated to -6° F. in the Cold Exchanger and expanded to 24.7 psia in the Expander. The expansion serves two purposes: it supplies energy to drive the Oxygen Compressor; it cools the expanded 90% hydrogen stream sufficiently to permit its use as a source of intermediate refrigeration in the plant.

The cold expander exhaust chills the purge gases from the cell, the cracked ammonia from the vaporizer, the condensate used for scrubbing ammonia out of the crude hydrogen, and the compressed crude hydrogen before it is scrubbed with water. After it has supplied this refrigeration, a side stream of it is used to carry off water vapor from the drier bed which is being regenerated. The side stream is returned to the main stream and the combined streams are heated to 150° F. and saturated with water

vapor in the Hydrogen Saturator. (This item is shown on our flowsheets but is considered outside our battery limits.) The warm saturated 90% hydrogen stream flows to the fuel cell battery at 14.7 psia.

The liquid oxygen feed is split into two streams. One supplies refrigeration at a low level (-328° F.) for producing reflux in the Scrubbing Column. The other supplies refrigeration at a somewhat higher level to meet the overall requirements of the cryogenic system.

The low level oxygen stream is vaporized at 1.4 psia in indirect heat exchange with the hydrogen-nitrogen feed to the scrubbing column. It is then heated to -6° F. in the Cold Exchanger and compressed to 16.7 psia in the Oxygen Compressor.

In parallel with the low level oxygen stream, the higher level stream is vaporized and heated in the Cold Exchanger at slightly above atmospheric pressure. After it leaves the Exchanger at -6° F., it re-joins the compressed low level stream. The combined streams are heated to 150° F. and saturated in the Oxygen Saturator. (The item, like the Hydrogen Saturator, is shown on our flowsheets but is considered outside our battery limits.) The warm saturated oxygen flows to the fuel cell battery at 14.7 psia.

Due to inert impurities in the feed gases, the fuel cells must be purged on both the hydrogen and oxygen sides. Since our hydrogen feed contains 10% nitrogen, the purge rate must exceed this flow. The oxygen feed contains 0.5% argon, and the oxygen side must be purged to remove this inert impurity.

One further criterion for determining the rate of purge is the allowable concentration of the inert impurities on each side. We have selected purge rates which keep these concentrations within the ranges

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permitted by the supplier of the fuel cell. In addition we have set the purge rate from the hydrogen side high enough to permit us to mix the hydrogen and oxygen purge streams without entering the explosive range.

As a result we can chill the mixed purge streams and safely compress the mixture to the suction pressure of the Crude Hydrogen Compressor. The argon, oxygen and nitrogen in these streams go to the Scrubbing Column where they tend to collect in the liquid nitrogen which goes to storage. Thus there is no gas to be vented during operation of this plant.

Equipment Description and Arrangement

Most of the equipment used in this process is conventional in design. Therefore, only the aspects which are unusual, or items which require special materials of construction, will be described. The arrangement was arrived at with the aid of the model shown in Items 12 and 13 in the Appendix. (The supports shown in the photographs are intended for assembly of the model only.)

The Ammonia Preheater utilizes inconel in its construction, due to high operating temperature and the possibilities of nitriding and hydrogen embrittlement. This is also true of the Ammonia Dissociator. Both units may be made of newer alloys now under investigation for resistance to these forms of attack. The Dissociator is a set of six fixed-bed catalytic reactors in parallel. In the design we are preparing, the catalyst is made of iron. Our own laboratory data indicates that certain other materials might permit a more compact design. This is a subject for investigation in the future program recommended by us.

The Crude Hydrogen Compressor, like the Oxygen Compressor and the Off-Gas Booster, is a non-lubricated, slow-speed, reciprocating compressor. Prospective vendors have assured us that the noise specification can be met. Lubrication is avoided in order to prevent carryover of oil or decomposition products which foul the low temperature equipment or cause fires in oxygen piping. Glass-filled teflon rings, or rings made of other materials, have been proved satisfactory in similar service.

The Cold Exchanger is a brazed-aluminum, extended-surface heat exchanger of a type which has been used widely in low temperature service. Its high surface per unit volume and high fin efficiency make it especially attractive in this application. Exchangers of this type have passed leak tests with freon and helium.

Originally the Scrubbing Column was planned to be a packed column. It became evident during design that the required enrichment could be obtained by differential partial condensation, which has often been utilized in cryogenic applications. The column is simply a vertical shell-and-tube heat exchanger with the nitrogen condensing out of the gas as it rises through the tubes. Oxygen boils at low pressure in the shell. This design is believed to be less sensitive to changes in the trim or diving angle of the submarine than any packed or tray-type column would be. (Incidentally, the oxygen supply should be passed through an adsorbent to remove contaminants before vaporization or the vaporizer should be purged periodically to nitrogen storage to prevent their accumulation in the oxygen-rich bath.)

The equipment has been piped and arranged as shown on Drawing P-1-D in the Appendix. Taking into account the need for access, insulation, support (not shown), etc., it is arranged to project somewhat beyond the limits set by the Bureau of Ships, as indicated by the broken lines. In the piping design made for this study, the short runs with frequent bends would appear to favor flexibility and minimize thrust problems. However, if and when we are asked to prepare a final design, we would then check the piping for stresses due to expansion, contraction, pulsation and shock. The effect of pipe size and configuration on propagation of noise would also be taken into account.

Utilities

Electricity and cooling water are the only utilities used by the process. The electrical loads may be summarized as follows:

<u>Equipment</u>	<u>Maximum Operating Load (KW)</u>
Dissociator	457.0
Crude Hydrogen Compressor	141.8
Crude Hydrogen Dryers	4.0
Off-Gas Booster	9.7
Condensate Pump	0.321
Fresh Water Recycle Pumps	0.926
Instrumentation	1.0
Lighting	<u>1.0</u>
Total	615.7

Assuming that the fuel cells consume 0.1 lb. of hydrogen per KWH, the above process load requires the production of 61.6 lbs./hr. The remaining 78.4 lbs./hr. of the hydrogen output of this process exceeds by 12% the 70 lbs./hr. of net real hydrogen required for an acceptable design.

Cooling water at 95° F. is available from the FW/SW Heat Exchanger and is distributed as shown on Service Flowsheet FS-2-D. The total requirement is 17.8 GPM. In addition, 24.8 GPM of hot (160° F.) water is used for saturating the gases going to the cells, and 0.4 GPM of condensate is used to scrub out the undissociated ammonia in the crude hydrogen. The latter stream is wasted to the sea as dilute aqua ammonia.

Instrumentation and Manpower

The proposed instrumentation is shown on Engineering Flowsheet FS-3-D. Our standard symbols are explained on Item 11 in the Appendix. The aim of this plan is to minimize manual operation without using excessive space for the controllers.

When an increase in power output is required, the operator raises the control point on the ammonia flow controller. This may also be done remotely from the Propulsion Control Cubicle. All of the remaining adjustments follow automatically, where it is necessary.

The temperature of the Dissociator tends to fall, so the power input increases to maintain it. A time delay is built into the relay to avoid drawing increased power before it is produced. During the resulting temperature dip the amount of undissociated ammonia increases, but the constant flow of scrubbing water is sufficient to remove it.

The bypass on the Crude Hydrogen Compressor closes to maintain suction pressure. The timing cycle on the Dryers will handle the maximum load.

As the temperature of the crude hydrogen at the cold end of the Cold Exchanger rises, more liquid oxygen at atmospheric pressure is admitted to the exchanger to supply the additional refrigeration requirement. The purity of the hydrogen produced tends to fall with increasing load, and this closes the bypass on the Oxygen Compressor.

The rate of feed of both gases to the battery increases. When this reduces the purity of the purge streams, the bypass on the Off-Gas Booster closes. The minor instrumentation operates similarly.

The residence time of hydrogen in the equipment and piping at full rate is about 50 seconds. All of the instrumentation will respond

in less time than this. On the oxygen side of the process, the time required to fill the shell of the Scrubbing Column is about 15 minutes. Therefore, the instrumentation is designed to maintain the maximum level, permitting a varying purity of hydrogen. The hydrogen concentration never falls below 90%, however. A manual by-pass is provided around the level control valve for quick startup.

The short response time on the hydrogen side implies a low holdup in the equipment and piping. If the hydrogen inventory were liberated in the submarine, it would amount to about 350 cubic feet, which is well below the value of 900 cubic feet which would create an explosive atmosphere.

The panel for control of this plant is a free-standing panel about six feet high by seven feet long by two feet deep. An additional foot of working space is required behind the panel. All of the instruments are the miniature type; pressure gauges are the miniline type. DC electrical instrumentation, with a standby battery in each case, is used for reliability and comparative insensitivity to shock and vibration.

In view of the instrumentation provided, only about one quarter of one man's full time per watch is required. His duties include logging readings every two hours, blowing down compressors where necessary, checking crankcase levels, checking calibration of analyzers, etc.

Cost Estimate

The installed cost of this plant would be \$800,000, including contingency. This is based on our experience in construction of chemical process plants. The figure includes design, engineering, purchasing, equipment, labor and supervision of construction by the engineering firm. Construction in a Navy shipyard is assumed. We have also assumed that all utilities are brought to the battery limits by others.

Spare parts are not included, since we are not familiar with Navy practice in this area. Demolition and removal of existing equipment is also excluded.

Assumptions

The major assumption made in designing this process is that the equipment can be mounted or re-designed where necessary to meet the noise and shock specifications. For ordinary heat exchangers, pumps, columns and compressors this does not seem to imply much change, if any. The Dissociator presents special problems.

Due to the high operating temperature of the Dissociator, its mechanical design cannot be based on high safety factors. This makes it vulnerable to added loads such as shock. Ordinary types of high-temperature insulation are not chosen for strength. Also high-capacity DC relays must close and open quickly and are likely to be noisy. We have assumed that all of these problems can be solved by improvements in materials and circuitry.

With reference to the power supply, we have assumed that a DC supply of constant voltage will be available. This contrasts with the drooping characteristic of the present chemical batteries. We believe that mechanical switching equipment can be designed to alter the series-parallel arrangement as the load changes, and keep the output voltage constant.

In general, we have assumed the following efficiencies: fuel cell 60%; compressors 70%; pumps 70%; motors 70% to 90% (depending on size). The design pressure of the oxygen storage tank is assumed to be equivalent to the static pressure due to 700 feet of sea water. The operating pressure of oxygen storage will be 16.7 psia. Liquid ammonia will be stored at 167 psia and 85° F.

Most of our detailed assumptions were design bases which have been verified in the course of design. The major exception to this is the efficiency

of the Scrubbing Column. We have assumed that this unit will approach true differential condensation in operation. If it does not, the reflux ratio, hence the quantity of oxygen to be compressed, will have to be increased.

Regarding the fuel cell itself we have conferred with General Electric, the supplier, and have made the following assumptions. The cell will operate at about 150° F. Feed gases must be saturated with water vapor at that temperature and 14.7 psia. Purge gases will also be saturated under the same conditions.

PROCESSES UTILIZING DISSOCIATION OF AMMONIA

Alternate II - Dissociation at High Pressure by Electrical Energy

Alternate II - Dissociation at High Pressure by Electrical Energy

General Description

This alternate is viewed as an intermediate stage in the development of a submarine hydrogen plant based on ammonia dissociation. It has not been studied intensively and will be described briefly with reference to Alternates I and III.

It differs from Alternate I chiefly in that the ammonia is cracked at, or near to, the pressure at which the dissociation products are separated. Thus, the greater part of the duty of the Crude Hydrogen Compressor is assumed by a much more compact liquid ammonia pump, which requires less power and makes less noise.

Although a high-pressure dissociator for this process is offered by at least one vendor, we are not aware of any industrial experience with this type of equipment. The nitriding and/or embrittlement which attack low-pressure dissociators would be more rapid at high pressure. Also the percentage of undissociated ammonia in the gas leaving the dissociator is expected to increase as the pressure is raised, if dissociation is limited by approach to equilibrium, as seems likely. It may be assumed that there will be little or no saving in dissociator volume. Also the cell off-gases will have to be compressed all the way to the pressure at which the dissociation products are separated, as in Alternate III. If the purge streams are compressed separately for additional safety, the compressors will occupy almost as much space as the Crude Hydrogen Compressor of Alternate I. It appears then that Alternate II offers little advantage in space over Alternate I.

Since it does consume less fuel cell power, however, the entire plant may be scaled down. To avoid detailed design calculations, the following approximations have been used;

1. The parasitic power load of Alternate I is reduced by deleting the requirements of the Crude Hydrogen Compressor and the Off-Gas Booster.
2. The requirements of the Off-Hydrogen and Off-Oxygen Boosters are added to the result to get a parasitic power load for Alternate II.
3. The hydrogen flow corresponding to this parasitic power is added to the design requirement of 70 lbs./hr. of net real hydrogen to get a total hydrogen output for Alternate II.
4. Using material balance ratios from Alternate I, the oxygen and ammonia requirements are calculated from the hydrogen output.
5. Finally, the plant volume is calculated from that of Alternate I by assuming volumes vary directly with hydrogen output.

PROCESSES UTILIZING DISSOCIATION OF AMMONIA

Alternate III - Dissociation at High Pressure by Chemical Energy

Alternate III

Dissociation of Ammonia at High Pressure by Chemical Energy

General Description

In this process the high parasitic power load of Alternate I is reduced in two ways: (1) combustion of ammonia in the process stream is used, instead of electrical energy, to supply the heat of dissociation; (2) dissociation is carried out at the pressure required to separate the products, as in Alternate II. At present we consider this the most efficient process that has been proposed. In view of its low power requirement, the combination of process and fuel cell should be more compact than that of any other process under consideration.

A firm design of this alternate would require bench scale and pilot development of a catalyst or catalysts and a dissociator. Since this design at present includes more uncertainty than Alternate I, we have not carried our material balances, energy balances and equipment designs as far as those of Alternate I. We have not sized or laid out the piping as in Alternate I, but the flows and conditions show that it would occupy less space in this case.

Process Description

The following description refers to Drawing No. FS-5-D (Item 7 in the Appendix to this report). As in Alternate I the process consists of dissociation of ammonia followed by condensation of the nitrogen produced. The hydrogen stream and the oxygen used in its purification are fed to the fuel cell.

Liquid ammonia is pumped to 340 psia in the Ammonia Pump. It is vaporized, preheated and fed to the Ammonia Dissociator. That part of the oxygen supply which is used for combustion is also pumped to separation pressure in the Liquid Oxygen Pump, vaporized, preheated and fed to the Ammonia Dissociator.

The dissociator is a two-stage, fixed-bed catalytic reactor. At the top of each bed the oxygen feed to that bed reacts completely with part of the ammonia and hydrogen, liberating heat which elevates the gas temperature. In the rest of the bed as the gas supplies the heat of dissociation of the remaining ammonia, the temperature falls again. This simplified two-step mechanism leads to the most conservative design.

On a dry basis the cracked ammonia contains about 0.1% residual ammonia and no oxygen. This stream is used to preheat ammonia and oxygen. On being cooled and finally chilled most of the water produced by combustion of ammonia condenses, along with the residual ammonia. The weak aqua ammonia drops out in the Separator and is wasted to the sea.

The crude hydrogen is dried and sent to the Cold Exchanger. Here it is cooled and partly condensed before going to the Scrubbing Column, which is actually a partial condenser, as in Alternate I. Liquid nitrogen is drawn off the bottom of the column. When the pressure is reduced to atmospheric, the hydrogen flashes off along with some nitrogen. The

remaining liquid nitrogen goes to storage in a vacant compartment of the liquid oxygen tank. The flash gas is re-compressed in the Off-Hydrogen Booster.

The overhead product of the Scrubbing Column is 90% hydrogen, 10% nitrogen, as in Alternate I. This is warmed to 26° F. in the Cold Exchanger and expanded to 24.7 psia. Here again it supplies power for the Oxygen Compressor and intermediate-level refrigeration for the plant before it is heated, saturated and fed to the fuel cells.

Two refrigerant oxygen streams are used, as in Alternate I. After compression of the low-level stream, they are combined, heated, saturated, and fed to the fuel cells.

In Alternate I the cell off-gases were mixed before recompression. This entailed an increase in the hydrogen purge to keep the oxygen concentration of the mixture outside the flammable range. We included backup instrumentation to insure safe operation.

In this alternate the purge gas streams are compressed separately and injected into the crude hydrogen ahead of the Dryers. Thus the purge streams can be sized for the minimum power requirement compatible with satisfactory operation of the fuel cell. The oxygen concentration is well below the flammable limit at all times. However, there is some loss of simplicity and compactness, which can be borne by this process.

Equipment Description and Arrangement

The arrangement of this equipment is shown on Drawing P-2-D (Item 9 in the Appendix). This indicates that the allowable space is exceeded, but by much less volume than Alternate I. There was insufficient time to study special designs of compressors; use of custom built units for these small loads might bring the total volume within the allowable space.

Here again much of the equipment is conventional and will not be described. Even liquid oxygen pumps, for example, are now in widespread use in industry and rocketry. The pressures in this case are not exceptional.

The Ammonia Dissociator is a completely new design. Since oxidation and dissociation are both carried out catalytically in this piece of equipment, there must be two catalysts or one catalyst that will accelerate both reactions. There is considerable published information on oxidation catalysts which indicates that the rate is high and that the side reaction which produces oxides of nitrogen may be minimized. The Research Laboratories of our company have developed data on dissociation catalysts which appear to be better at low pressure and low conversions than those in commercial use and to be effective in the presence of water vapor. These may also catalyze the oxidation reaction. Application of the above mentioned data to our fixed-bed reactor model has given us a preliminary reactor design.

This is a down flow reactor with two fixed beds in series. The shell is lined with refractory which can expand independently on heating. Tentatively the shell is made of carbon steel; if hot spots develop, it may be necessary to use alloys to reduce nitriding and embrittlement. The space between beds for injection of oxygen is filled with material to reduce the risk of explosion.

Although the Off-Gas Boosters are shown as separate machines, they are actually a multipurpose compressor driven by one motor. Incidentally, this is also true of the Crude Hydrogen Compressor and Off-Gas Booster of Alternate I. Machines of this type minimize space and maintenance requirements and are widely used in industry.

Utilities

The most attractive characteristic of this Alternate is its low power consumption:

<u>Equipment</u>	<u>Maximum Operating Load (KW)</u>
Liquid Oxygen Pump	0.165
Ammonia Pump	0.718
Crude Hydrogen Dryers	4.000
Off-Hydrogen Booster	19.300
Off-Oxygen Booster	0.718
H ₂ Saturator Bottoms Pump	0.773
O ₂ Saturator Bottoms Pump	0.347
Instrumentation	1.000
Lighting	<u>1.000</u>
Total	28.02

Assuming again that the fuel cells consume 0.1 lb. of hydrogen per KWH, the above process load requires the production of 2.80 lbs./hr. The remaining 98.0 lbs./hr. of the usable hydrogen output of this process exceeds by 40% the 70 lbs./hr. of net real hydrogen required for an acceptable design.

The water distribution is shown on Process Flowsheet FS-5-D (Item 7 in the Appendix). The cooling water requirement is 26.3 GPM at 95° F. In addition, 2.7 GPM of 160° F. water is used for saturating the gases going to the cells; and 0.28 GPM of condensate from the process is wasted to the sea as dilute aqua ammonia.

Instrumentation and Manpower

The proposed instrumentation is shown on Engineering Flowsheet FS-4-D. As before, the control point to be re-set with change in load is the ammonia flow rate. The remaining adjustments follow automatically.

In this case, the temperatures in the Dissociator are maintained by changes in the flow rate of oxygen to the unit and to the bottom bed. Otherwise, the major process instrumentation follows the pattern of Alternate I.

Exact residence time has not been calculated, since we have not firmly sized the equipment or piping. It is less than the 50 seconds calculated for Alternate I, largely because the Dissociator is smaller. For the same reason the hydrogen inventory is less than 350 cubic feet. The panel dimensions, instrumentation types, and operator manpower are the same as those given for Alternate I above.

Cost Estimate

The installed cost of this plant would be \$515,000, including contingency. The bases of this estimate are the same as those for Alternate I, with one addition. We have assumed the success of a development program without estimating its cost. Under Future Program we have outlined the development work involved; details will be furnished on request.

Assumptions

All of the assumptions listed for Alternate I, except those for the Dissociator, apply to Alternate III. We have indicated the process assumptions for the Dissociator under Equipment Description and Arrangement: available low pressure and low conversion data may be applied to operation at high pressure and high conversion; a single catalyst will accelerate both reactions or the added volume for a separate oxidation catalyst is negligible; under the chosen conditions the extent of the side reactions leading to oxides of nitrogen is negligible.

The mechanical design limitations for this Dissociator are much less stringent than those for the Dissociator of Alternate I. Here the design temperature of the pressure vessel is far below the reaction temperature, since the process heat is not transferred through it. This allows a higher working strength and reduces the risk of attack by the process gases. The insulation may be supported on the lower head of the shell without being tied to it, thus reducing the transmittal of shock loads to it. Therefore, the assumption that this unit can be designed to meet the shock test is less questionable than it is for Alternate I.

FUTURE PROGRAM

FUTURE PROGRAMS

Generation of hydrogen for submarine propulsion by decomposition of ammonia is indicated by this study to be feasible, using any one of the three processes discussed in this report. These processes have several items of equipment and instrumentation in common. Therefore, it is possible to design, procure and test the common equipment while other items are being developed. Using this possibility should enable us to reduce the time required to test the fuel cell installation.

The following programs to begin at the same time are suggested. Program No. 2 is considered marginal and may be dropped without affecting performance of Program No. 3, which is considered the most promising one.

1. Begin firm design of the equipment to supply hydrogen by Alternate I. Procure and test the individual items on a test bed and modify where necessary to withstand shock and minimize noise.
2. Design, build and operate a pilot dissociator for Alternate II. If this is demonstrated successfully before program 1 is completed, design and purchase a full scale pressure dissociator. Re-design associated process equipment only where necessary, converting the plant design to Alternate II.
3. Begin bench scale and pilot plant studies of the dissociator of Alternate III. Depending on the time required this program would fit in with the previous ones as follows:

- a. If piloting is completed successfully before programs 1 and 2 are completed, the plant design should be changed where necessary to convert it to Alternate III.
- b. If piloting is completed successfully after Alternate I or II is demonstrated, the installed plant should be modified to Alternate III.

The following subjects would be investigated in the laboratory and pilot plant:

1. Effect of changes in pressure, temperature and concentration on catalyst activity.
2. Relative rates of ammonia cracking vs. ammonia-hydrogen oxidation.
3. Catalyst life and mechanical integrity under the high pressure and temperature conditions.
4. Materials of construction and dissociator design details.
5. Explosive limits in the ammonia-hydrogen-oxygen-water vapor system under the operating conditions.

Details of the development program are available if required.

BIBLIOGRAPHY

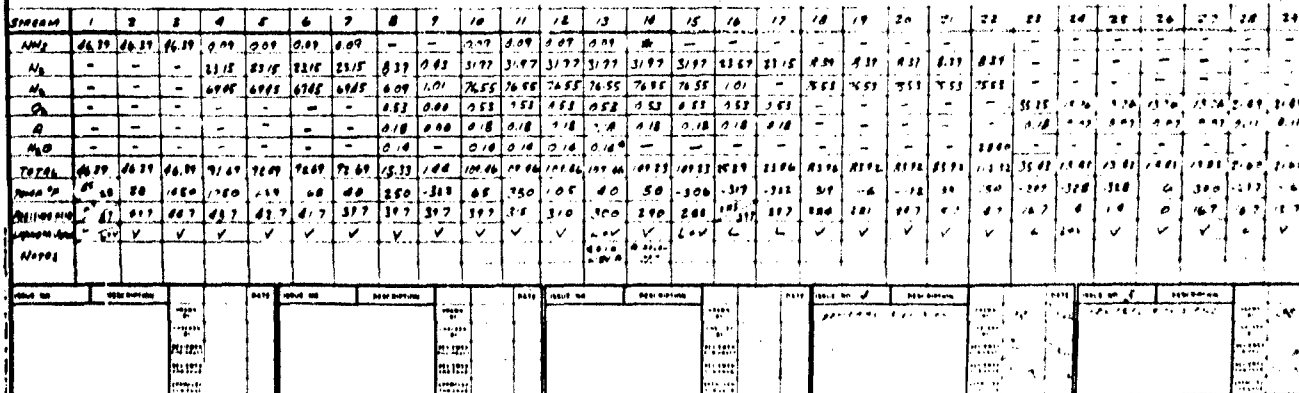
BIBLIOGRAPHY

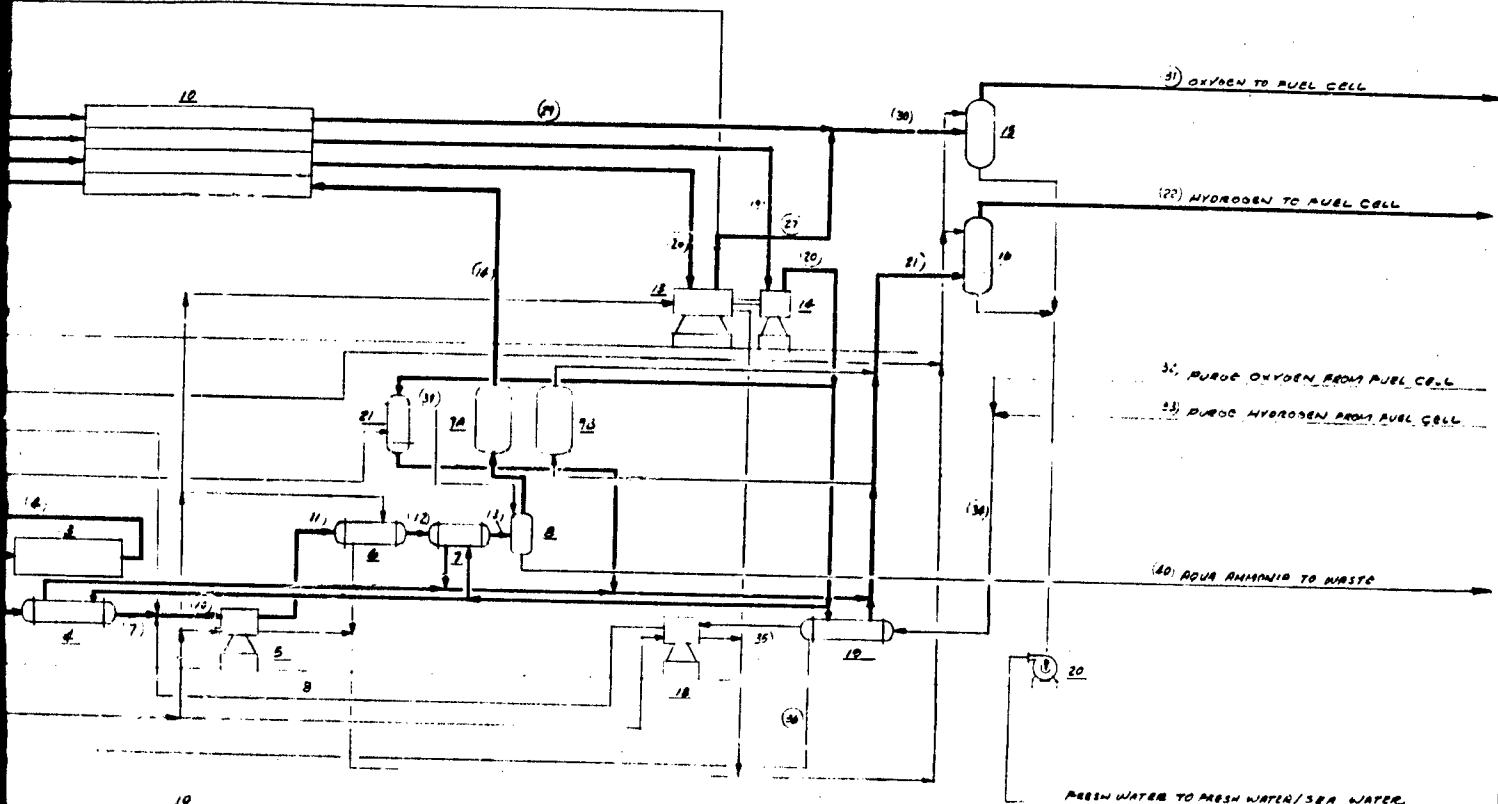
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APPENDIX

APPENDIX

<u>Item</u>		<u>Drawing No.</u>
1	Electrical Energy for Dissociation - Process Flowsheet Alternate I	FS-1-D
2	Electrical Energy for Dissociation - Service Flowsheet Alternate I	FS-2-D
3	Electrical Energy for Dissociation - Engineering Flowsheet Alternate I	FS-3-D
4	Line Schedule for Engineering Flowsheet FS-3-D Alternate I	LS-1-B
5	Line Schedule for Engineering Flowsheet FS-3-D Alternate I	LS-2-B
6	Chemical Energy for Dissociation - Engineering Flowsheet Alternate III	FS-4-D
7	Chemical Energy for Dissociation - Process Flowsheet Alternate III	FS-5-D
8	Electrical Energy for Dissociation - General Arrangement of Equipment and Piping - Alternate I	P-1-D
9	Chemical Energy for Dissociation - General Arrangement of Equipment - Alternate III	P-2-D
10	Summary of Process Requirements and Outputs	
11	Drafting Standards and Symbols for Flow Diagrams	FS-6-D
12	Photograph of Model of Alternate I Installation (facing port)	
13	Photograph of Model of Alternate I Installation (facing starboard)	




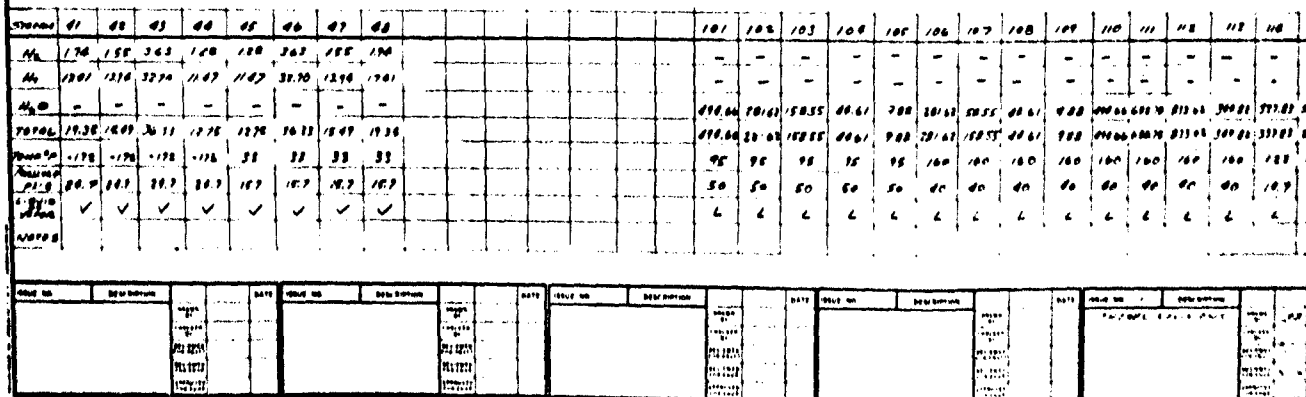
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NOTES

- 1) ALL PLANTS ARE AN OUTPUT OF TENDING OF REAL NUTRIENTS ABOUT THE AMOUNT REQUIRED FOR POWER TO RUN THE OPERATION AND PURIFICATION PROCESS
- 2) ALL PLANTS ARE IN BALANCE HOLDING
- 3) -25 WATT PLANT ENERGY FOR GREEN WATER PLANT ENTERS AND ASSOCIATED PROCESS PLANT

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FORM NO.	ISSUANCE	REV. DATE	REV. REASON	DATE
			LIST OF PLOTS	
				
<p>GENERATION OF HYDROPOWER BY DISIPATING RUN OF RIVERS IN INTERSTATE & ELECTRICAL ENERGY AND DISSEM. OF POWER TO THE PEOPLE. (EYANGAR CORPORATION) (EYANGAR PROJECT) (EYANGAR PROJECT) (EYANGAR PROJECT)</p>				
<p>DATE</p>		<p>PROJECT NO. (EYANGAR PROJECT)</p>		<p>DATE</p>
<p>1. NAME OF THE PROJECT (EYANGAR PROJECT)</p> <p>2. NAME OF THE PROJECT (EYANGAR PROJECT)</p> <p>3. NAME OF THE PROJECT (EYANGAR PROJECT)</p> <p>4. NAME OF THE PROJECT (EYANGAR PROJECT)</p> <p>5. NAME OF THE PROJECT (EYANGAR PROJECT)</p> <p>6. NAME OF THE PROJECT (EYANGAR PROJECT)</p> <p>7. NAME OF THE PROJECT (EYANGAR PROJECT)</p> <p>8. NAME OF THE PROJECT (EYANGAR PROJECT)</p> <p>9. NAME OF THE PROJECT (EYANGAR PROJECT)</p> <p>10. NAME OF THE PROJECT (EYANGAR PROJECT)</p>		<p>EXTENDING TO THE</p> <p>EXTENDING TO THE</p>		<p>DATE</p>
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1) ALL FLOWS ARE FOR OUTPUT OF 70 LB/HR OF 25% HYDROGEN ABOVE THE AMOUNT REQUIRED FOR POWER TO RUN THE GENERATION AND PURIFICATION APPARUS

2) ALL FLOWS ARE IN POUNDS UNLESS NOTED

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<p><i>Specified for use in accordance with the following:</i></p> <p>ALUMINUM 2. ELECTRICAL ENERGY AND DISTRIBUTION SPARKING PLUGS SWIFT 13.54011 CAN PROTECT THIS 100 PRODUCT SERIAL 48-12 01 00 OF PAGE 0010</p>			
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PROJECT ENGINEERING

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ALQ-1	NH ₃ STORE.	NH ₃ VAPORIZER	3"	S/S	NO	-	SCH. 5S						
AV-1	NH ₃ VAPORIZER	NH ₃ PREHTR.	2"	S/S	YES		SCH. 5S						
AV-2	NH ₃ PREHTR.	NH ₃ DISSOCIATOR	3"	S/S	YES								
HNA-1	NH ₃ DISSOCIATOR	NH ₃ PREHTR.	4"	S/S	YES								
HNA-2	NH ₃ PREHTR.	NH ₃ VAPORIZER	3"	S/S	YES		SCH. 5S						
HNA-3	NH ₃ VAPORIZER	CRUDE H ₂ COOLER	3"	S/S	NO		SCH. 5S						
HNA-4	CRUDE H ₂ COOLER	LINE HI-2	3"	S/S	YES		SCH. 5S						
OF-1	OFF-GAS BOOSTER	CRUDE HYD. COOLER	2"	S/S	YES		SCH. 10S						
HI-1	FLASH TANK	LINE HI-2	3"	AL	YES		SCH. 40						
HI-2	LINE HNA-4	CRUDE H ₂ COMP.	3"	S/S	YES		SCH. 5S						
HI-3	CRUDE H ₂ COMP.	CRUDE H ₂ AFTER-COOLER	2"	S/S	YES		SCH. 10S						
HI-4	CRUDE H ₂ AFTER-COOLER	CRUDE H ₂ CHILLER	2"	S/S	NO		SCH. 10S						
HI-5	CRUDE H ₂ CHILLER	WATER SCRUBBER	2"	S/S	YES		SCH. 10S						
HI-6	WATER SCRUBBER	COLD EXCH.	2"	S/S	YES		SCH. 10S						
HI-7	COLD EXCH.	SCRUBBING COLUMN	1"	AL	YES		SCH. 40						
NIL-1	SCRUBBING COLUMN	FLASH TANK	1"	AL	YES		SCH. 40						
NIL-2	FLASH TANK	LIO. NITROGEN STORAGE	1"	AL	YES		SCH. 40						
H-1	SCRUBBING COLUMN	COLD EXCH.	1"	AL	YES		SCH. 40						
H-2	COLD EXCH.	EXPANDER	2"	S/S	YES		SCH. 10S						
H-3	EXPANDER	LINES H-2, H-3, H-4 (DWG. LS-2-B)	2"	AL	YES		SCH. 40						
H-4	H ₂ SATURATOR	H ₂ SATURATOR	3"	S/S	NO		SCH. 5S						
H-5	H ₂ SATURATOR	FUEL CELL	4"	S/S	YES		SCH. 5S						
LOX-1	LIQUID OXYGEN STORAGE	LINES LOX-2, LOX-3	1"	AL	YES		SCH. 40						
LOX-2	LINE LOX-1	SCRUBBING TOWER	2"	AL	YES		SCH. 40						
O-1	SCRUBBING TOWER	COLD EXCH.	3"	AL	YES		SCH. 40						
O-2	COLD EXCH.	OXYGEN COMP.	4"	S/S	YES		SCH. 5S						
O-3	OXY COMP.	LINE O-5	2"	S/S	YES		SCH. 10S						
LOX-3	LINE LOX-1	COLD EXCH.	1"	AL	YES		SCH. 40						
O-4	COLD EXCH.	LINE O-5	2"	S/S	YES		SCH. 5S						
O-5	LINES O-3 & O-4	OXYGEN SATURATOR	3"	S/S	YES		SCH. 5S						
O-6	OXYGEN SATURATOR	FUEL CELL	4"	S/S	YES		SCH. 5S						
O-7	FUEL CELL	LINE OF-2	1"	S/S	YES		SCH. 5S						
HI-8	FUEL CELL	LINE OF-2	2"	S/S	YES		SCH. 5S						
OF-2	LINES O-7 & HI-8	OFF-GAS CHILLER	2"	S/S	YES		SCH. 5S						
OF-3	OFF-GAS CHILLER	OFF-GAS BOOSTER	2"	S/S	YES		SCH. 5S						

ISSUE	P	D											
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DRAWN BY	ST												
APPROVED BY													
DATE	7/15/64	7/20/64											

AMERICAN CYANAMID COMPANY
ENGINEERING & CONSTRUCTION DIVISION
NEW YORK, N.Y., U.S.A.

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S ENGINEERING

SPG	VISC	SERVICE FACTOR	OVER DESIGN FACTOR	CORROSION ALLOWANCE	LINE DATA										STREAM NO	SERVICE CLASS NO
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ANCE USED IN DETERMINATION OF PIPE WALL THICKNESS SHALL BE AS FOLLOWS
 FERRITIC ALLOYS (LOW ALLOY STEEL) 0.030" SHALL BE USED.
 LESS STEEL ALLOYS, 0.030" SHALL BE USED
 REQUIRE HIGHER CORROSION ALLOWANCE IF SO NOTE THEM UNDER THIS COLUMN

LINE SCHEDULE FO

AUTHORIZATION NO 1250-244	DRAWING NO LS-1-B
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LINE DATA										STREAM NO	SERVICE CLASS NO	REMARKS
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WALL THICKNESS SHALL BE AS FOLLOWS
 0.050" SHALL BE USED.
 IF SO NOTE THEM UNDER THIS COLUMN

LINE SCHEDULE FOR FLOW SHEET

AUTHORIZATION NO. 120-244	DRAWING NO. LS-1-B	REF FLOW SHEET NO. 120-3-D
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1

450

AMERICAN CYANAMID COMPANY
ENGINEERING & CONSTRUCTION DIVISION
NEW YORK, N.Y., U.S.A.

LINE SCHEDULE FOR
FLOW SHEET

REFERENCES

PS-1-D, PS-2-D
PS-3-D

1-MINIMUM CORROSION ALLOWANCE USED IN DETERMINATION OF PIPE W
(a) FOR CARBON STEEL AND FERRITIC ALLOYS (LOW ALLOY STEEL) 0.005
(b) FOR AUSTENITIC OR STAINLESS STEEL ALLOYS, 0.030" SHALL BE US
2-SPECIAL CONDITIONS MAY REQUIRE HIGHER CORROSION ALLOWANCE IF

[illegible]

LINE SCHEDULE FOR
FL

AUTHORIZATION NO PCO-244	DRAWING NO LS-7-B	REV CS
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4

LINE DATA

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- FS-1-D

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ALL THICKNESS SHALL BE AS FOLLOWS
 30" SHALL BE USED.
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 SO NOTE THEM UNDER THIS COLUMN

LINE SCHEDULE FOR
 FLOW SHEET

AUTHORIZATION NO

PEO-244

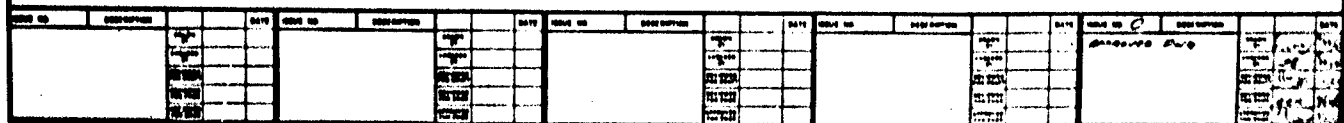
DRAWING NO

LS-2-B

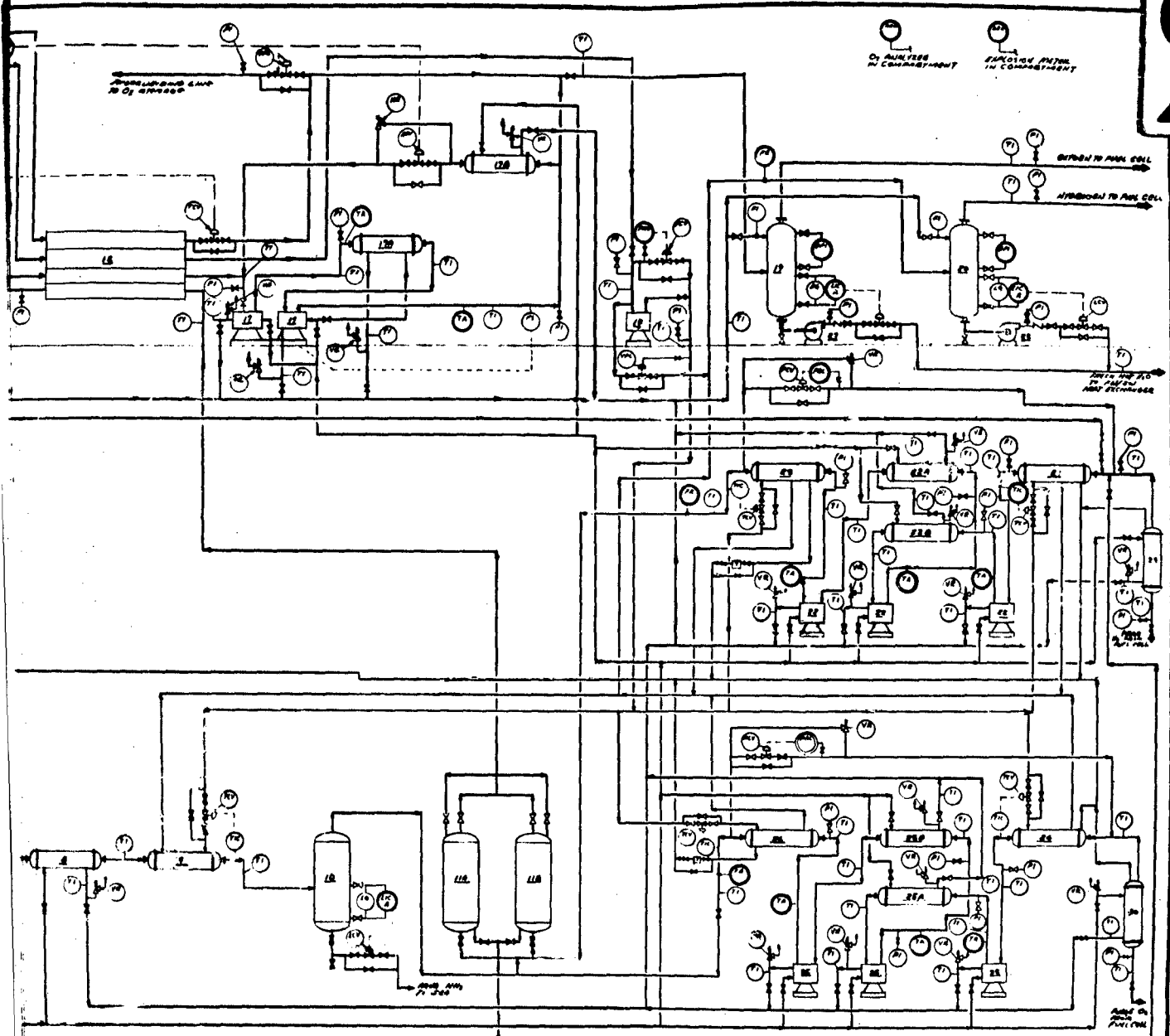
REF FLOW SHEET NO

FS-1-D, FS-2-D
FS-3-D

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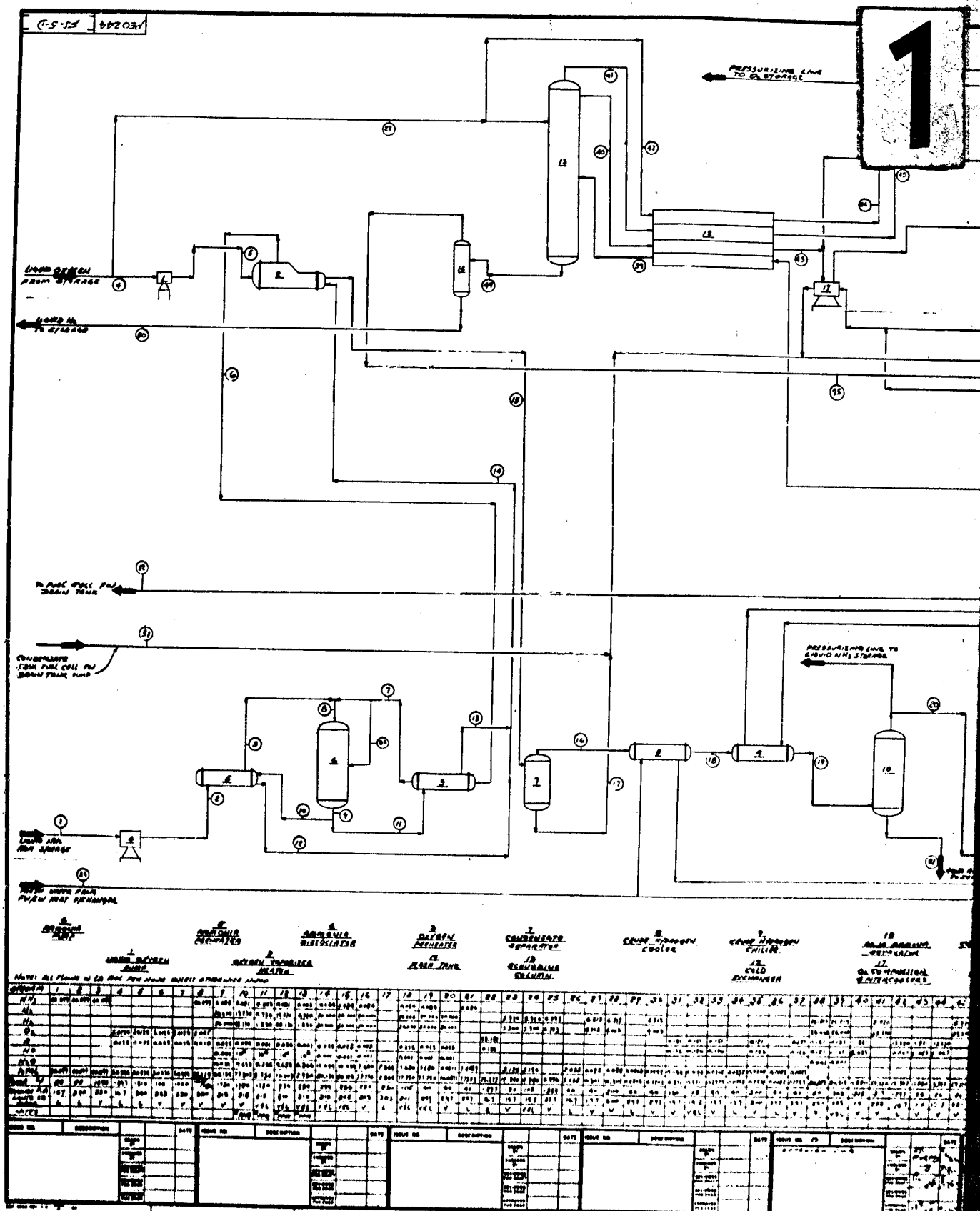


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NOTE: PUMP SIZES, IN GENERAL, ARE SMALL TO BE SMALLER THAN IN ALTERNATE E. PUMP SIZES ARE SMALLER THAN IN ALTERNATE E. SEE DRAWINGS 13-1-B AND 13-1-C

CYANAMIDE	
SEPARATION OF HYDROGEN BY DISSOCIATION OF AMMONIA	
APPROXIMATE CHEMICAL ENERGY FOR DISSOCIATION	
APPROXIMATE FLOW SHEET: US NAVY CONTRACT NO. 13-1-B	
PROJECT SERIAL NO. 13-1-B AND 13-1-C	
DATE	DATE
DESIGNED BY	DESIGNED BY
CHECKED BY	CHECKED BY
APPROVED BY	APPROVED BY
140-210	15-4-D

REVISION	DATE	DESCRIPTION	REVISION	DATE	DESCRIPTION
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CONFIDENTIAL
CONFIDENTIAL

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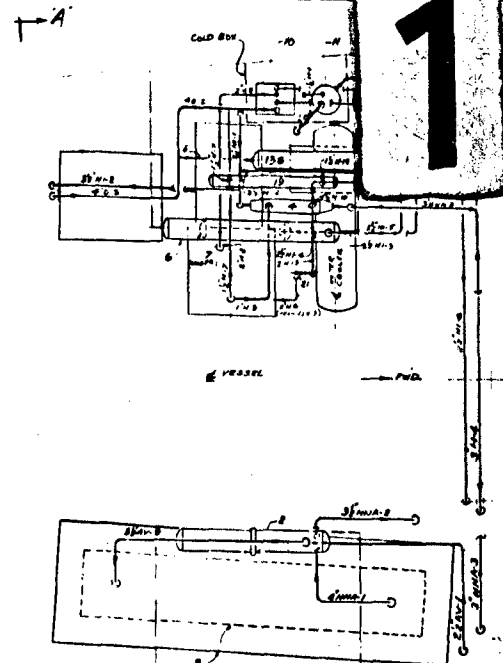
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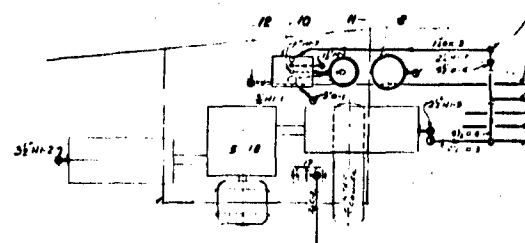
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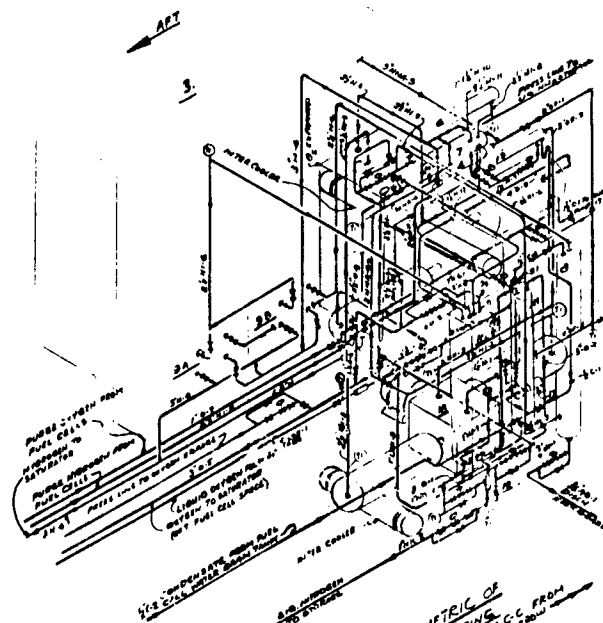


PLAN VIEW U.P.
UPPER LEVEL

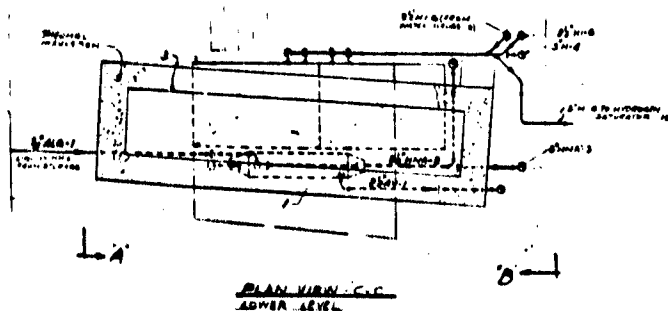


PLAN VIEW C.C.
CONCRETE LEVEL

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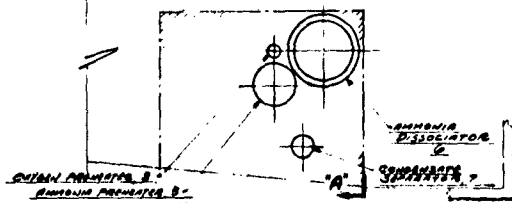
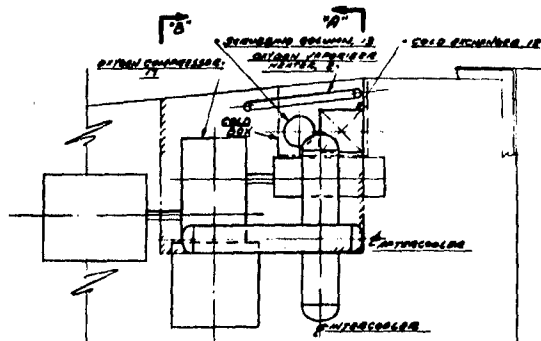


PLAN VIEW - C.C.
ADWIA - 454

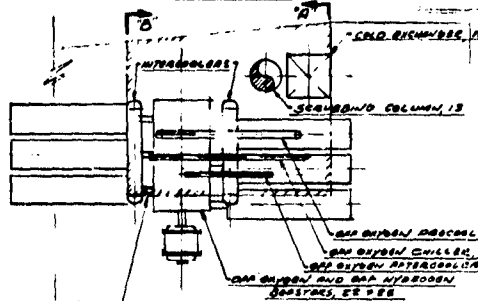
1. ALTERNATE 1, ELECTRICAL ENERGY FOR DISSOCIATION
FUEL COST = \$1.00
2. ALTERNATE 1, ELECTRICAL ENERGY FOR DISSOCIATION
SERVICE FUEL COST = \$1.00
3. ALTERNATE 2, ELECTRICAL ENERGY FOR DISSOCIATION
FUEL COST = \$1.00
4. PUMP LINE SCHEDULE FOR SWAGE PUMP LINE \$1.00
\$1.00
5. PUMP LINE SCHEDULE FOR SWAGE PUMP LINE \$1.00
\$1.00
6. DRAFTING STANDARDS AND SYMBOLS FOR \$1.00
7. SURVEYING - PHOTOGRAPHIC SURVEY AND PHOTOGRAPHY
FUEL COST IN \$1.00 WILL BE \$1.00

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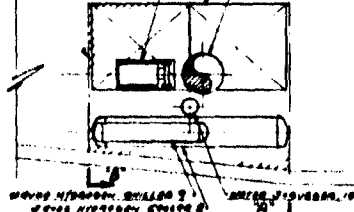
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18" PLAN VIEW C.C.
UPPER LEVEL



DOT AND DASH LINE INDICATE SERVICE
LIMITS ALLOCATED BY U.S. NAVY.



PLAN VIAN O.D
JANUARY 1966

EXHIBIT 2009

1. ALTERNATE EL. CHEMICAL ENERGY FOR DISSOCIATION.
SUBMISSIONS FLOW SHEET ----- TWO PG. 6-D
2. ALTERNATE EL. CHEMICAL ENERGY FOR DISSOCIATION.
PROCESS FLOW SHEET ----- TWO PG. 8-D
3. ENGINEERING - FEASIBILITY STUDY: HYDROGEN OXIDE FUEL CELL
IN STEEL HULL AEROBIC SUBMERGENT. BY DR. JAMES
M. HARRIS. JAN 1978. 10 PG. 10-D

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ITEM 10
SUMMARY OF PROCESS REQUIREMENTS & OUTPUTS

<u>Alternate</u>	<u>I</u>	<u>II</u>	<u>III</u>
Usable Hydrogen Produced, lbs/hr. (Note A)	140.0	118.4	100.8
Parasitic Power Load, KW	615.8	484.3	28.0
Net Real Hydrogen Produced, lbs/hr. (Note B)	78.4	70.0	98.0
Ammonia Usage (lbs. NH_3 /lb Net Real H_2)	10.1	9.54	6.96
Oxygen Usage (lbs. O_2 /lb Net Real H_2)	14.5	13.6	9.9
Total Equipment Weight, tons (estimated)	60	-----	25
Gross Plant Volume, cu. ft.	1480	1252	1295
Fuel Cell Volume for Parasitic Power, cu. ft. (Note C)	205.2	161.4	9.3
Cooling Water Requirement, GPM	17.8	-----	26.3
Estimated Start-up Time, hrs.	8	8	2
Additional Load to Air Conditioning Refrigeration Requirements, Tons	11	-----	0
Fuel Requirements, BTU/Net KWH			
Full Load	8081	-----	5568
1/2 Load	12062	-----	5732
1/4 Load	20569	-----	6090
Fuel Costs, ¢/Net KWH (Note D)			
Full Load	4.04	-----	2.78
1/2 Load	5.06	-----	2.86
1/4 Load	10.28	-----	3.04
Oxygen Costs, ¢/Net KWH (Note D)			
Full Load	4.35	-----	2.97
1/2 Load	5.44	-----	3.05
1/4 Load	11.07	-----	3.24

- Notes: A. "Usable hydrogen" is real (100%) hydrogen fed to the cell less real hydrogen purged from the cell.
 B. "Net real hydrogen" is usable hydrogen less hydrogen required for the parasitic power load, assuming 0.1 lbs. of real hydrogen required per KWH produced by the cell.
 C. This assumes 3 KW/cubic foot of cell volume.
 D. Fuel and Oxygen Costs are based on \$80.00/ton ammonia and \$60.00/ton oxygen.

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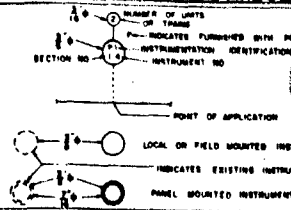
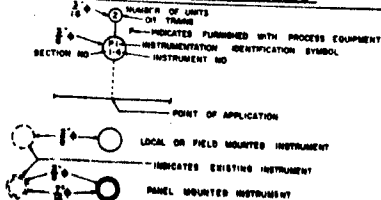
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TABLE II
INSTRUMENT SYMBOLS

























ANALYSIS		44	ANALYSIS RECEPTOR
45	ANALYSIS ALARM	45	ANALYSIS RECEPTOR CONTROLLER
46	ANALYSIS CONTROLLER	46	ANALYSIS TRANSMITTER
47	ANALYSIS CONTROLLER	47	ANALYSIS TRANSMITTER
48	ANALYSIS INDICATOR		
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151	CONDUCTIVITY CONTROLLER		
152	CONDUCTIVITY CONTROLLER	</	

TABLE III
GRAPHIC SYMBOLS FOR USE ON FLOW DIAGRAMS

	ANGLE VALVE (HIGH PRESSURE TYPE)		HOSE, FLEXIBLE (ARMORED)
	ANGLE VALVE (LOW PRESSURE TYPE)		HOSE, METAL (FLEXIBLE)
	AIR CYLINDER OPERATED VALVE		HOSE, PLASTIC
	SSOO W/P VALVE (NEEDLE TYPE)		HOSE, RUBBER
	SPOO W/P VALVE (DISC TYPE)		HYDRAULIC CYLINDER OPERATED VALVE
	BLOCK VALVE NEEDLE TYPE 3 WAY (ELECT OPERATED)		LEVEL GLASS
	BLOW OFF OR BLOW DOWN VALVE (FAUCET VALVE)		LINE SEPARATOR
	BURNER		LINE BLIND VALVE (WASHER TYPE)
	BUTTERFLY OR SLAPPER VALVE		NEEDLE VALVE
	PILOT TUBE		BAROMETRIC CONDENSER
	CHECK VALVE (LIFT OR SWING)		OPEN DRAIN
	CHECK VALVE (DIP BLOCK TYPE FOR DOWN)		PISTON OPERATED VALVE
	CHOKE VALVE (PISTON OPERATED)		POPPIN VALVE
	CHECK VALVE (BALL TYPE)		PLUG VALVE
	CHECK VALVE (RECESS PLUG TYPE)		PLUG OR COCK VALVE (THREE WAY)
	DIAPHRAGM VALVE PNEUMATIC OPERATED		PLUG OR COCK VALVE (FOUR WAY)
	DIAPHRAGM VALVE (SELF OPERATED)		RESTRICTION ORIFICE
	DIAPHRAGM MOTOR OPERATED VALVE		QUENCH VALVE
	DIAPHRAGM VALVE (BLOCK) (SOLENOID) (PNT) (SERVICE)		QUICK OPENING VALVE (LEVER OPERATED)
	DIAPHRAGM VALVE (SOLENOID) (PNT) (SERVICE)		
	DRINKING FOUNTAIN		REDUCER (CONCENTRIC)
	EYE WASH FOUNTAIN		REDUCER (ECCENTRIC)
	EJECTOR		REPURGE SOL. IN USE (OR VACUUM BREAKER)
	ELECTRIC MOTOR OPERATED VALVE		RELIEF VALVE (SAFETY)
	EXHAUST HEAD		
	EXPANSION JOINT (COMMERCIAL, BELLOWS TYPE)		SEPARATOR
	EXPANSION JOINT (COMMERCIAL, SLIDE TYPE)		SHOWER HEAD
	EXPANSION JOINT (SPECIAL DESIGN FOR BELLOWS TYPE)		SIGHT GLASS
	EXPANSION JOINT (SPECIAL DESIGN FOR SLIDE TYPE)		SOLENOID VALVE
	EXPANSION VALVE		STEAM TRAP
	FIGURE SIGHT		STOP CHECK VALVE
	FLAME ARRESTER		STRAINER (TEMPORARY)
	FLAME OR TOUCH		STRAINER Y TYPE WITH VALVE
	FLOW ELEMENT OR ORIFICE		STRAINER Y TYPE WITH PLUG
	FLOW TOTALIZER (DISPLAY UNIT)		STRAIGHT THROUGH VALVE
	FOOT VALVE		VENTURI OR FLOW METER
	FOUL STOP		CONDENSATE VENT
	GATE VALVE		INTERNAL TANK VALVE
	GATE VALVE (JACKETED)		WEIGHT OPERATED VALVE (GATE)
	GLOBE VALVE		WEIGHT OPERATED VALVE (GLOBE)
	BLOCK VALVE (ISOLANT)		VENTURI AIR LINE (OR GAS VENTURI)
	HIGH OR LOW PRESSURE (OR GAS) VALVE		VENTURI WITH ELECTRICAL (FLOW, OR GAS, WEIGHT)
	HIGH OR LOW PRESSURE (OR GAS) VALVE		VENTURI WITH CAP (LOW PRESSURE, OR GAS, WEIGHT)

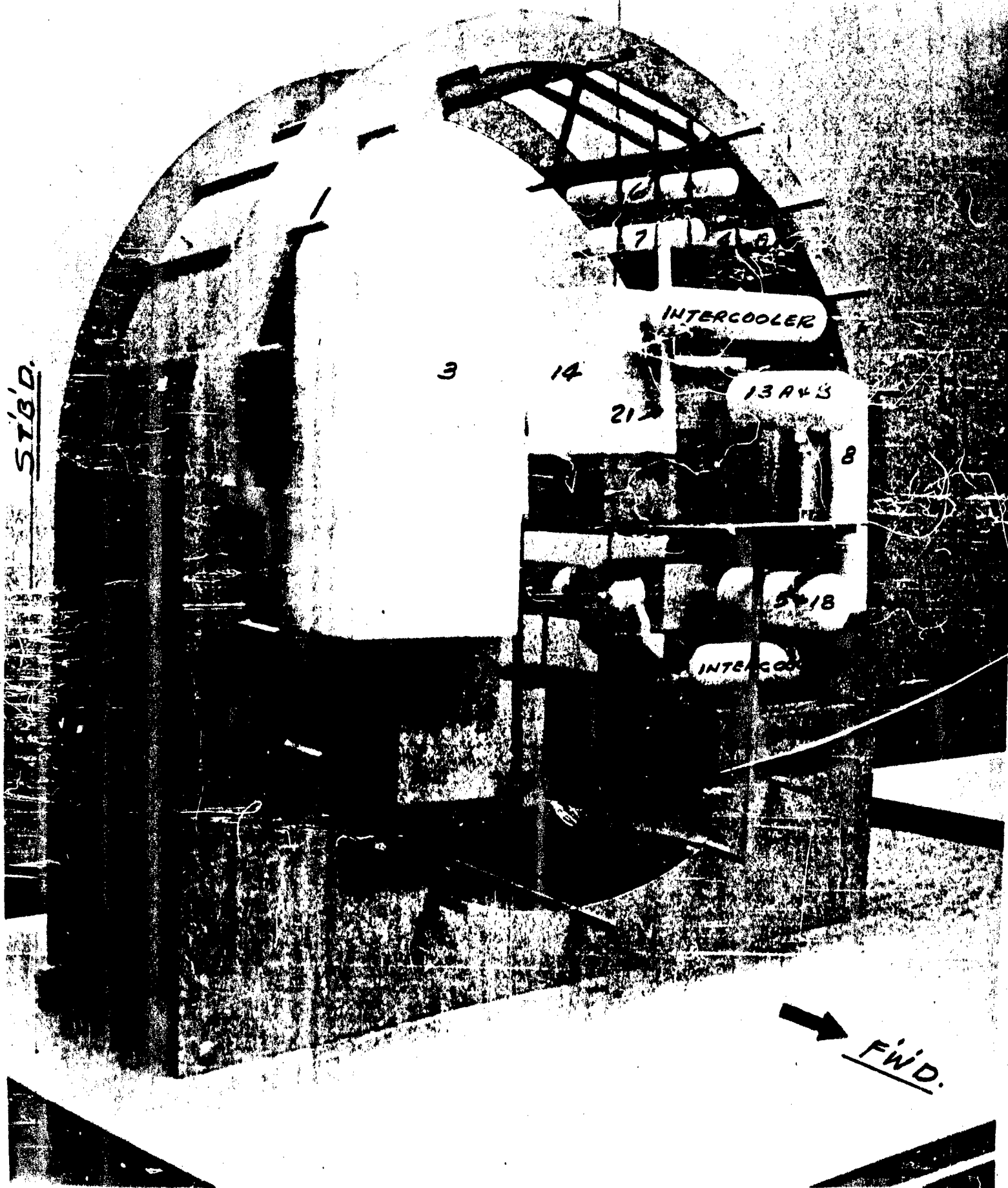
TABLE IV
EQUIPMENT SYMBOLS (PROCESS)

	CENTRIFUGAL PUMP		ROTARY OR GEAR PUMP
	RECIPROCATING PUMP		VACUUM PUMP
	PROPORTIONING PUMP		SUMP PUMP
	DOUBLE STAGE RECIPROCATING COMPRESSOR		MULTIPLE OR TWO STAGE RECIPROCATING COMPRESSOR
	CENTRIFUGAL COMPRESSOR		ROTARY COMPRESSOR
	THREE STAGE COMPRESSOR		VERTICAL PUMP
	VERTICAL PUMP		SEPARATOR
	TANK WITH HEATING COIL		AIR TANK CAR
	BOX TYPE CONDENSER OR COOLER		STORAGE TANK
	CENTRIFUGAL FAN		HEAT EXCHANGER
	HEAT EXCHANGER OR COOLER		GLOBE VALVE

**DRAWING STANDARDS & SYMBOLS
FOR FLOW DIAGRAMS**

FIG. 1

STBD.

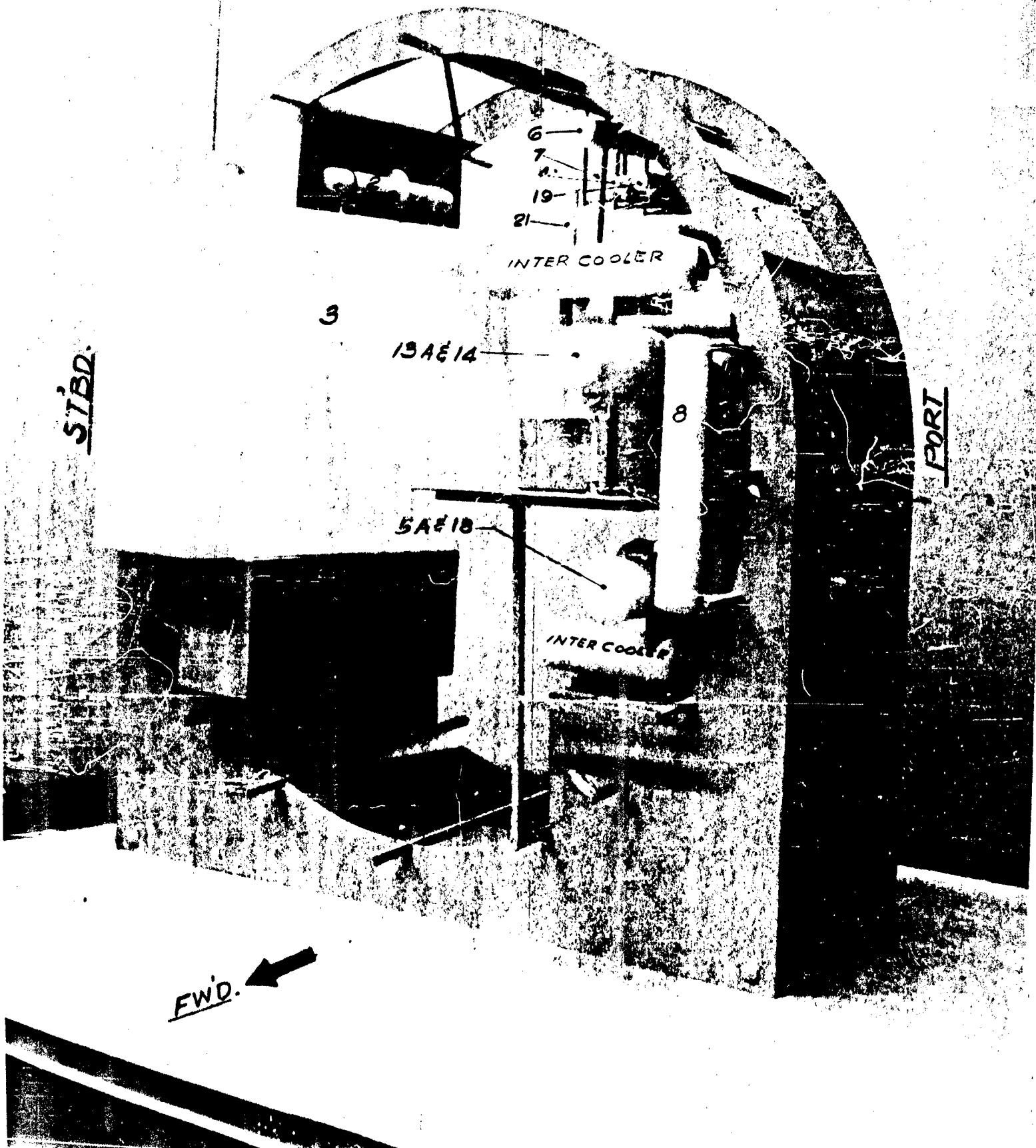


MODEL VIEW - ALTERNATE I.
LOOKING TO PORT

P20-344
P-4-B

3	AMMONIA DISSOCIATOR
4	CRUDE HYDROGEN COOLER
5A	CRUDE HYDROGEN COMPRESSOR & INTERCOOLER 5B
6	CRUDE HYDROGEN AFTER COOLER
7	CRUDE HYDROGEN CHILLER
8	WATER SCRUBBER
9A	CRUDE HYDROGEN DRYER
9B	CRUDE HYDROGEN DRYER
11	SCRUBBING COLUMN
13A	OXYGEN COMPRESSOR, INTERCOOLER & RECYCLE COOLER 13-B
14	EXPANDER
18	OFF GAS BOOSTER
19	OFF GASS CHILLER
21	CONDENSATE CHILLER

PEO-244
P-3-B.



MODEL VIEW - ALTERNATE I
LOOKING TO STBD

UNCLASSIFIED

UNCLASSIFIED